Convective Wall-to-Suspension Heat Transfer in Circulating Fluidized Bed Risers

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A mechanistic model, which incorporates recent findings on the fluid dynamics in the riser of the circulating fluidized bed (CFB), is developed for predicting the suspension-to-wall heat-transfer coefficient in the riser. It is assumed that heat transfer between the gas-particle suspension and the riser wall takes place by the contact of both particle packets and an emulsion phase on the wall. A characteristic length (L), that is, a sliding distance of the emulsion phase along the heat-transfer surface, is introduced in the model, enabling the effect of the length of heat-transfer surface to be evaluated. It is found that the heat-transfer coefficient decreases with increasing L, but becomes increasingly insensitive to L when L is larger than 1 m. Agreement between model prediction and measurement is encouraging over a range of operating conditions, heat-transfer surface length, and riser diameters.

Introduction

In recent years circulating fluidized beds (CFBs) have been extensively used for coal combustion and gas–solid reactions. In the design of CFB boilers or reactors it is important that the heat-transfer coefficient be estimated with reasonable accuracy. Although heat transfer in CFB risers has received considerable attention in the literature, our understanding of the heat-transfer mechanism in a CFB is still incomplete and there is a degree of contradiction among reported experimental results. The influence of riser fluid dynamics on wall-to-suspension heat transfer in a CFB has been widely reported. Wu et al. (1989a) found that although the time-averaged heat-transfer coefficient increased linearly with the suspension density, the instantaneous heat-transfer coefficient could depend on measurement position if the suspension density was very low. By measuring instantaneous heat-transfer coefficient and capacitance signals, Wu et al. (1991) found that the contact of solid packets on the heat-transfer surface corresponded to the instantaneous increase in the heat-transfer coefficient. By measuring radial temperature and solids flux profiles, Leckner and Andersson (1992) suggested that suspension-to-wall heat transfer may be due to the motion of particle packets near the heat transfer surface. Basu (1990) suggested that the heat transfer in a CFB riser may be due to the contact of renewable particle packets and the gas phase containing dispersed particles on the wall. The "core-annulus" hydrodynamic structure was first introduced by Wu et al. (1990) into heat-transfer modeling in a CFB riser. Mahalingam and Kolar (1991) proposed an "emulsion-layer" model for wall heat transfer in a CFB. The model proposed by Chen and Chen (1992) was based on the packet model of Mickley and Fairbanks (1955), but the "packets" used in the model were in fact individual particles in contact with the riser wall. In this article a new approach is applied to model the process of heat transfer in CFB risers. It assumes that the contact of packets on the riser wall is transient, while the emulsion phase sweeps along the wall in riser axial direction. This physical model is based on observation of the motion of solids near the riser wall. The model presented here enables the effect of the length of heat-transfer surface to be successfully predicted.

Experimental Studies

The CFB apparatus used in the present investigation incorporated a 0.152-m-ID riser, 6 m in height. Details of the apparatus are reported elsewhere (Rhodes et al., 1992b). The

Table 1. Properties of Particles

Mean Dia.: 75 μm	Density: 2,456 $kg \cdot m^{-3}$
$\epsilon_{mf} = 0.44$	$U_t = 0.40\ m \cdot s^{-1}$
Size Range (μm)	Under Size (%)
125–150	100
90–125	99.2
63–90	78.8
38–63	28.7
0–38	4.8

exit of the riser tapered conically into a 0.1-m-ID bend connected to the inlet of the primary cyclone. The solids exit from this cyclone was connected to the top of a 0.305-m-ID solids reservoir via a pneumatically operated diverter that was able to divert the entire flow rate of solids into a measuring bed. The flow of solids from the reservoir to the riser was controlled by a 0.1-m-ID L-valve. The main air feed to the riser was delivered by a Rootes-type blower and passed through a distributor of glass beads at the bottom of the riser. The powder used was FRF5, a nonporous alumina with surface-volume mean particle size 75 μm and particle density 2,456 $kg \cdot m^{-3}$. Details of physical properties of this powder are summarized in Table 1. With this experimental arrangement it was possible to independently adjust superficial gas velocity and imposed solids flux to achieve a wide range of suspension densities in the measurement zone. Mean suspension densities were deduced from pressure gradients in the riser, ignoring contributions of friction and acceleration.

Measurements of wall-to-suspension heat-transfer coefficients were made over a 60-mm-long aluminum riser section, mounted at height 2.6 m above the L-valve inlet and heated from the outside using a flexible electrical heater. Guard heaters were employed to eliminate heat losses from the main heater. The output of guard heaters was adjusted until the temperature gradient across the lagging around the main heater was zero, indicating no net heat flow from the main heater to the surroundings. This ensures that we know the energy input accurately and that heat is transferred radially through the wall in the measurement section. Thermocouples of 10-ms response time and 1-mm OD were used to measure the inner and outer surface temperatures of the lagging, the inner surface temperature of the heat-transfer surface, and the temperature of gas–solid suspensions. Details of the apparatus and experimental procedures for heat-transfer measurements are described in Zhou (1992).

Heat-Transfer Modeling

A physical model

Measurement of the motion of solids and gas in the region near the riser wall is very difficult, but it is now possible to piece together a reasonable picture by combining the results of observations from several sources and using various measurement techniques. From previous studies (e.g., Bierl et al., 1980; Hartge et al., 1988; Rhodes et al., 1988; Weinstein et al., 1986) we know that the flow of the gas-particle suspensions in a riser is characterized by a rapidly rising, dilute core surrounded by a slowly falling, denser region adjacent to the walls. The work of Rhodes et al. (1992b) confirmed the observation of Monceaux and coworkers (1986) that there exists a similar-profiles regime, in which the reduced solids flux (local solids flux divided by cross-sectional mean solids flux) is insensitive to change in imposed solids flux over a wide range for a given superficial gas velocity. They have shown that the onset of the similar-profiles regime takes place at a higher mean solids flux as the superficial gas velocity is increased. The work of Rhodes et al. (1992a) using high-speed video demonstrated that the downward velocity of packets or swarms of particles at the wall was independent of superficial gas velocity and mean solids flux. These video studies suggested that packets at the wall were formed from particles ejected from the core and that the packets were eventually destroyed by the action of pulses of gas from the core. Their experimental results also suggested that the characteristics of packets such as the probability of appearance and the mean contact time on the wall was closely related to the onset of the similar-profiles regime.

Based on the riser fluid dynamics discussed earlier, we propose the following physical model (shown in Figure 1) for heat transfer from gas-particle suspension to the riser wall. Particles from the core region are thrown into the wall where

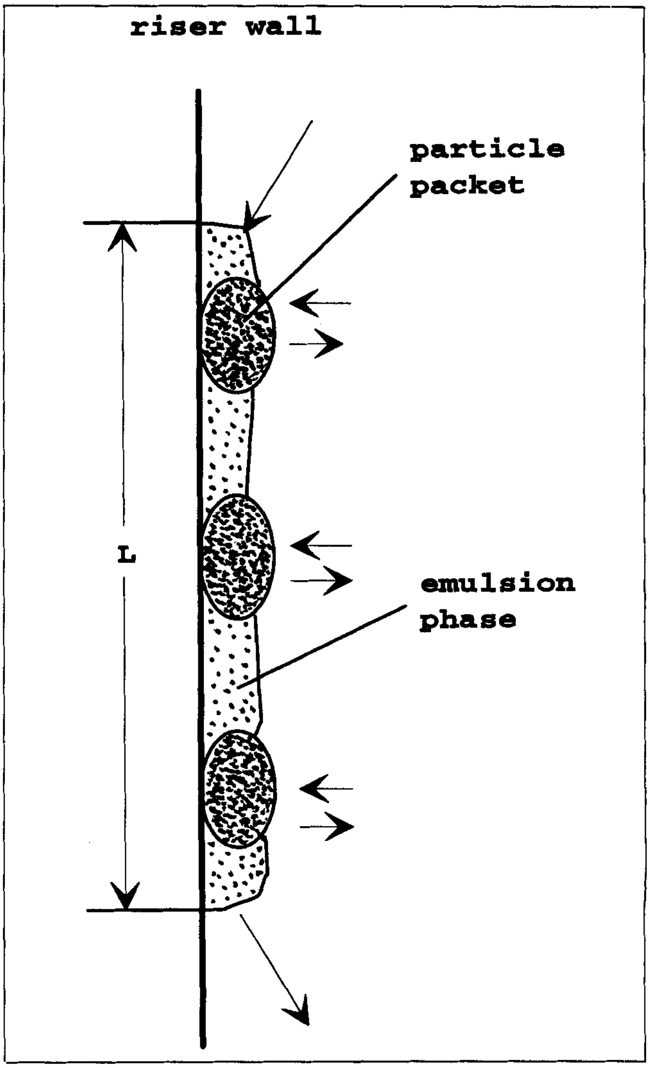

Figure 1. Physical model of suspension-to-wall heat transfer.

they form packets under the action of wall friction and gravity. After a random residence time at the wall these packets are picked up by the gas in the boundary region near the riser wall. Some of the packets are destroyed while others may bounce at the wall and return to the core region. An emulsion layer (or emulsion phase) is formed near the wall due to wall effect and destruction of particle packets on the wall. This emulsion layer slides downward along the riser wall for a distance L before being destroyed by turbulence in the riser. Some packets may come into contact with the wall by penetrating the emulsion layer.

It can therefore be envisaged that heat transfer between the gas-particle suspension and the riser wall is due to the renewing contact of the packets and the emulsion phase on the wall. The time-averaged heat-transfer coefficient may be expressed as

$$h = (1-\alpha) h_{el} + \alpha h_{pt}, \tag{1}$$

where α is the time-averaged fraction of heat-transfer surface in contact with the packets, and h_{pt} and h_{el} are time-averaged convective heat-transfer coefficients due to the packets and the emulsion phase, respectively. From Eq. 1 we see that this model does not take into account the contribution of radiative heat transfer, and so it may only be valid at low temperatures (e.g., under 400°C). It is shown by Zhou (1992) that it is the suspension density rather than the superficial gas velocity or the imposed solids flux that independently affects the heat transfer in CFB risers. The discussion in this article will therefore focus on the effect of the suspension density on the overall heat-transfer coefficient.

Heat transfer due to the emulsion phase

The emulsion-phase heat-transfer coefficient may be estimated using the Farag and Tsai (1993) approach, which is based on the transient gas-convective heating of single particles when sliding over the heating plate. The emulsion-phase heat-transfer coefficient is expressed as

$$h_{el} = \{d_p \rho_p C_p/\theta\} \cdot \{1 - \exp[-(12 k_g \theta)/(d_p C_p d_p^2)]\} \cdot \{K \rho_{el}/(\rho_{el} + K_s \rho_p d_p^3)\}, \tag{2}$$

where K and K_s are constants, and ρ_{el} is the suspension density of the emulsion phase near the riser wall. According to Rhodes et al. (1992c) ρ_{el} can be estimated from:

$$\rho_{el} = 2 \rho_{\text{susp}}, \tag{3}$$

where ρ_{susp} is the cross-sectional mean suspension density in the riser. θ in Eq. 2 is particle-surface contact time, expressed as (Farag and Tsai, 1993)

$$\theta = L/U_p, \tag{4}$$

where L is the length of the heat-transfer surface in contact with the emulsion phase, defined by Farag and Tsai (1993) as the characteristic length of the heat-transfer surface. The introduction of L in the present model enables the effect of heat-transfer surface length on heat-transfer coefficients to be predicted. In Eq. (4) U_p is the mean velocity of the emulsion phase. Since we are dealing with a fairly dilute suspension (i.e., the emulsion phase) in the region of the riser where gas velocity is negligible, we may assume that U_p is equal to the terminal velocity of a single particle.

Heat transfer due to solid packets

The unsteady-state heat-transfer equation for packets in contact with a heat-transfer surface was solved by Mickley and Fairbanks (1955); this equation gives:

$$h_{p1} = \sqrt{\frac{k_{pt} \rho_{pt} C_p}{\pi}} \int_0^\infty \tau^{-1/2} \varphi(\tau)\, d\tau, \tag{5}$$

where k_{pt} and ρ_{pt} are the effective thermal conductivity and density of packets, respectively. According to Gelperin and Einstein (1971), the thermal conductivity of packets may be calculated from:

$$k_{pt} = k_g \left[1 + \frac{(1-\epsilon_{pt})(1 - k_g/k_p)}{k_g/k_p + 0.28 \epsilon_{pt}^{0.63(k_p/k_g)^{0.18}}} \right], \tag{6}$$

where ϵ_{pt} is the voidage of packets, which is approximately equal to the voidage of the powder at minimum fluidization, that is, 0.44 for the FRF5 powder used. The density of packets in Eq. 5, ρ_{pt}, may be estimated from:

$$\rho_{pt} = (1 - \epsilon_{pt}) \rho_p + \epsilon_{pt} \rho_g. \tag{7}$$

According to Rhodes et al. (1992a), the contact time distribution of packets on the wall may be expressed in the form:

$$\varphi(\tau) = \frac{1}{\tau_m} \exp\left(-\frac{\tau}{\tau_m} \right) \tag{8}$$

where $\varphi(\tau)$ and τ_m are the distribution function of contact time and the mean contact time of packets on the riser wall, respectively. The mean contact time of packets on the wall increases with increasing suspension density before becoming constant at a suspension density of around 40 $kg \cdot m^{-3}$ (Rhodes et al., 1992a).

Since the Mickley and Fairbanks model overpredicts the heat-transfer coefficient for short contact time of packets on the wall, it is suggested by Decker and Glicksman (1981) that there may be an effective gas gap between the packets and the heat-transfer surface. The heat-transfer coefficient through the gas gap between the packet and the wall may be expressed as

$$h_{p2} = \frac{k_g}{d_p/n}, \tag{9}$$

where d_p/n is the thickness of the effective gas gap. Value n is normally in the range 4 to 10 for bubbling beds, and $n = 6$ is recommended for design purposes (Xavier and Davidson, 1985).

The heat-transfer coefficient due to the packet can then be written as

$$h_{pt} = \frac{1}{1/h_{p1} + 1/h_{p2}}. \tag{10}$$

Time fraction of packets in contact with the riser wall

In order to estimate the overall heat-transfer coefficient (using Eq. 1), we now need to know the time-averaged fraction of the heat-transfer surface in contact with particle packets (α). As first approximation, it is assumed that α is proportional to the probability of appearance of packets (p_{pt}), that is,

$$\alpha = k_f \times p_{pt}, \tag{11}$$

where k_f is a contact coefficient of packets on the wall. Rhodes et al. (1992a) gives the relationship between p_{pt} and ρ_{susp}.

Model Verification

The unknown parameters in the model just discussed are the thickness of the effective gas gap between packets and the heat-transfer surface (d_p/n) and the contact coefficient of packets on the riser wall (k_f). By testing model prediction with measurement from various sources, we find that the combination of $n = 6$ and $k_f = 0.2$ enables the model to give the most promising predictions. It is interesting to note that value $n = 6$ is the same as that recommended by Xavier and Davidson (1985). The value $k_f = 0.2$ also appears reasonable since it yields realistic time-averaged fraction of riser wall in contact with packets. For example, when the suspension density was increased from 3.4 to 28.0 $\text{kg}\cdot\text{m}^{-3}$, the value α (calculated from Eq. 11) increased from 0.065 to 0.168, which is in the range observed by Soong et al. (1993) under similar conditions. The values of n and k_f may be independent of riser diameter, since they are governed by the characteristics of the downward-flowing layer adjacent to the wall that is mainly a wall effect (Rhodes et al., 1992b). We therefore use $n = 6$ and $k_f = 0.2$ in all the following comparisons of model predictions with measurements.

Comparison of model prediction with measurement in the 0.152-m-ID riser is shown in Figure 2, and it can be noted that model prediction agrees reasonably well with measurement. It is interesting to note from model prediction that the increase in the heat-transfer coefficient with suspension density for $\rho_{\text{susp}} < 10\ \text{kg}\cdot\text{m}^{-3}$ (which is probably prior to the onset of the similar-profiles region) is more significant than that in the higher suspension density region, for example, $\rho_{\text{susp}} > 10\ \text{kg}\cdot\text{m}^{-3}$ (i.e., in the similar-profiles regime). This is characteristic of the heat transfer in CFB risers. The "linear" relationship between the heat-transfer coefficient and the suspension density or the square-root of the suspension density often reported in the literature may therefore be an oversimplification, especially when the suspension density is very low.

The comparison of model prediction with measurement of Wu et al. (1989b) is shown in Figure 3. The measurement of Wu et al. (1989b) contains the heat-transfer coefficients at three axial positions in a 0.152-m-ID riser for 171-μm sand particles. In the absence of a mechanistic model, Wu et al. (1989b) suggested that when the suspension density was above 40 $\text{kg}\cdot\text{m}^{-3}$, the relationship between the heat-transfer coefficient and the suspension density was independent of measurement position, while when the suspension density was below 40 $\text{kg}\cdot\text{m}^{-3}$, this relationship varied with the measurement position. Our model prediction suggests, however, that this relationship may be independent of measurement position over the whole range of suspension density examined. The apparent scattering of the measured heat-transfer coeffi-

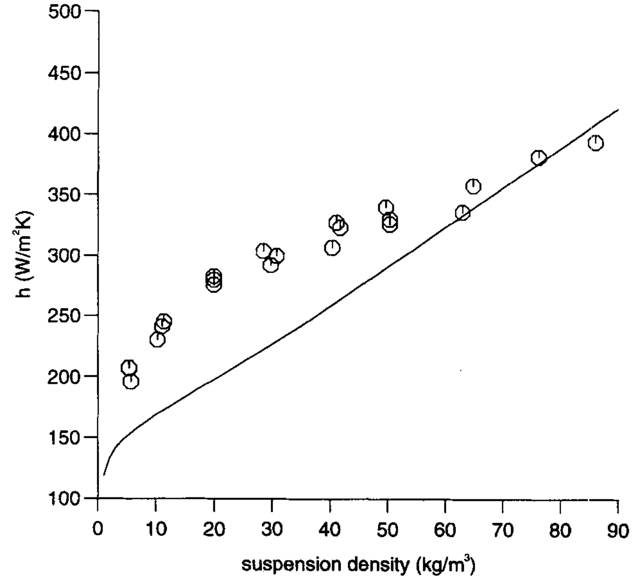

Figure 2. Comparison of predicted heat-transfer coefficient with measurement ($d_p = 75\ \mu$m).

Solid line: model prediction; box marker: measurement; riser diameter: 0.152 m. $L = 0.06$ m.

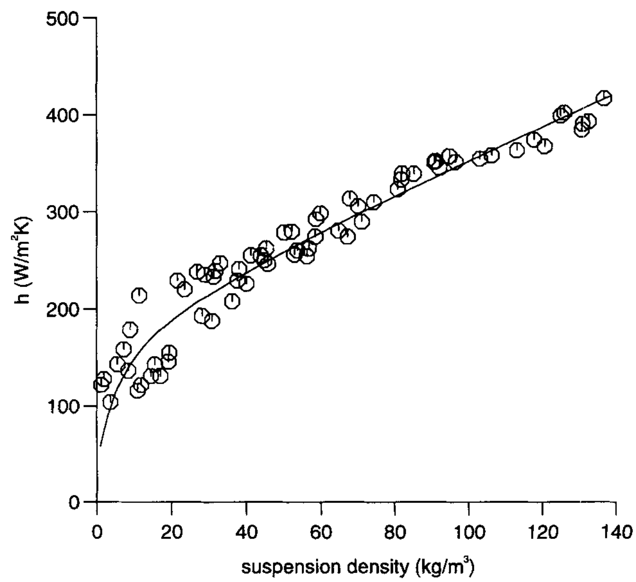

Figure 3. Comparison of predicted heat-transfer coefficient with measurement of Wu et al. (1989b) ($d_p = 171\ \mu$m).

Solid line: model prediction; box marker: measurement; riser diameter: 0.152 m. $L = 0.01$ m.

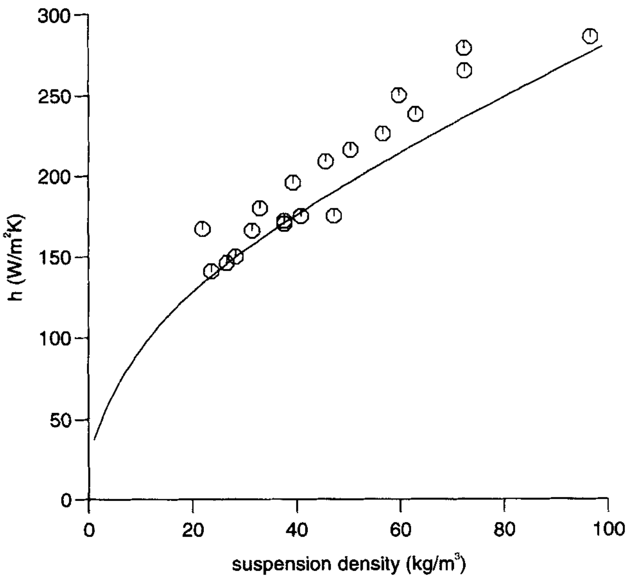

Figure 4. Comparison of predicted heat-transfer coefficient with measurement of Basu and Nag (1987) ($d_p = 227$ μm).

Solid line: model prediction; box marker: measurement; riser diameter: 0.102 m. $L = 0.01$ m.

cient at a suspension density of around 20 $kg \cdot m^{-3}$ may be due to the onset of the similar-profiles regime.

Model prediction also gives reasonable agreement with the data of Basu and Nag (1987) for the heat transfer in a 0.102-m-ID riser for 227-μm sand particles (Figure 4).

Figure 5 shows a comparison of model prediction with results of Kobro and Brereton (1986), in which an industrial-scale boiler was used. It can be noted that the heat-transfer coefficient increased almost linearly with suspension density.

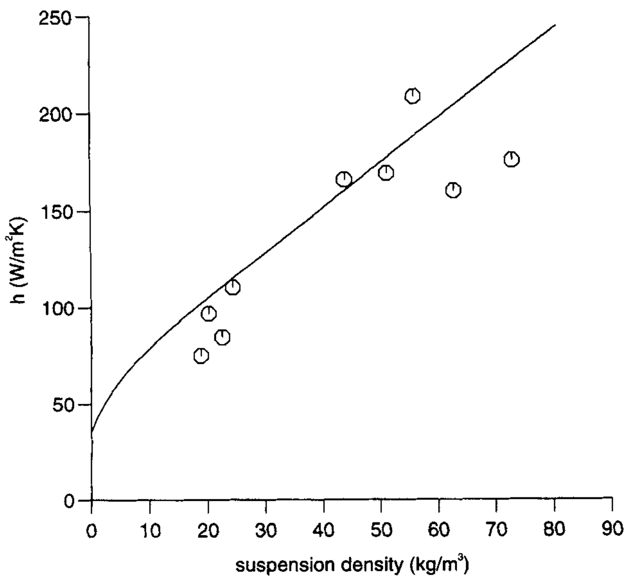

Figure 5. Comparison of predicted heat-transfer coefficient with measurement of Kobro and Brereton (1986) ($d_p = 170$ μm).

Solid line: model prediction; box marker: measurement.

Analysis of model prediction suggests that this "linear" relationship between the heat-transfer coefficient and the suspension density is due to the significant decrease in the contribution of the emulsion-phase heat transfer.

Effect of the Length of Heat-Transfer Surface

A few authors (e.g., Basu, 1990; Wu et al., 1990) have attempted to explain the reason for the decrease in heat-transfer coefficient with increasing length of the heat-transfer surface (i.e., so-called "length effect"). Their interpretation is based on the estimation that the mean contact time of packets on the heat-transfer surface increases with increasing length of the heat-transfer surface. The estimated packet contact time on the riser wall of Basu (1990) and Wu et al. (1990) are high compared to those used in this study. However, little information is available concerning the contact time of packets on the riser wall.

This model takes into account the "length effect" by introducing a characteristic length (L) of the heat-transfer surface. An example of the variation of model prediction with L is shown in Figure 6. It can be seen that the predicted heat-transfer coefficient decreases significantly with increasing L when L is very small (e.g., smaller than 0.01 m) and becomes insensitive to L when L is reasonably large (e.g., larger than 1 m). This agrees with the generally reported trend in literature. Analysis suggests that as L increases, the contribution of the emulsion phase heat-transfer decreases. For instance, at a suspension density 30 $kg \cdot m^{-3}$, as L increases from 0.001 to 0.01, then to 0.1 and then to 1.0 m, the contribution of the heat transfer due to the emulsion phase decreases from 67.2% to 59.8%, then to 21.9%, and then to 2.7%, respectively.

How do we select an appropriate value of L? For short and "straight" heat-transfer surfaces (e.g., Wu et al., 1989b), L is probably equal to the actual length of the heat-transfer surface. When the heat-transfer surface is long (e.g., Kobro and Brereton, 1986b), L may be smaller than the actual length of the heat-transfer surface. In this case the actual length of the heat-transfer surface may still be used in place of L for

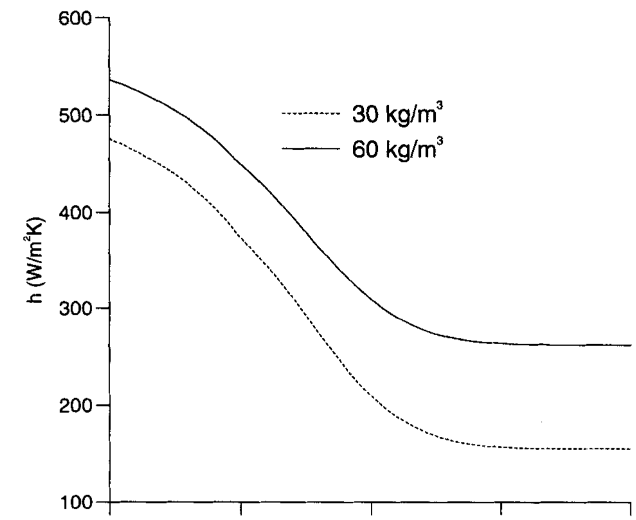

Figure 6. Effect of L on model prediction ($d_p = 75$ μm).

model prediction since it would yield almost the same result (see Figure 6). If the heat-transfer surface is not "straight" (e.g., Basu and Nag, 1987), then an equivalent length of the heat-transfer surface should be used.

It should be noted that the time fraction of heat-transfer surface in contact with packets used in the model represents a statistical mean of the random contact time of packets on the riser wall. Since any point of the riser wall would, in theory, be subjected to transient contact of packets, the model presented earlier would enable the prediction of time-averaged heat-transfer coefficients even for extremely small probes (e.g., $L = 0.001$ m).

Conclusions

A mechanistic model that incorporates up-to-date knowledge of riser fluid dynamics is proposed for predicting the convective heat-transfer in CFB risers. The proposed mechanism involves heat transfer by contact with the wall of both particle packets and an emulsion phase. The observed effect of the length of the heat-transfer surface on the heat-transfer coefficient is successfully modeled. The proposed model is capable of providing realistic predictions of the wall-to-suspension heat-transfer coefficient over a wide range of operating conditions. Analysis of model prediction suggests that the reported region of rapid increase in the heat-transfer coefficient with suspension density corresponds to a region prior to the onset of the similar-profiles region. Model predictions also compare reasonably well with the experimental data from several sources over a wide range of operating conditions.

Acknowledgments

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Notation

C_p = heat capacity of particles, $J \cdot kg^{-1} \cdot K^{-1}$
d_p = particle diameter, m
g = gravitational acceleration, $m \cdot s^{-2}$
h = overall heat-transfer-coefficient, $W \cdot m^{-2} \cdot K^{-1}$
h_{p1} = heat-transfer-coefficient of packets, $W \cdot m^{-2} \cdot K^{-1}$
h_{p2} = heat-transfer-coefficient through an effective gas film, $W \cdot m^{-2} \cdot K^{-1}$
k_g = thermal conductivity of gas, $W \cdot m^{-1} \cdot K^{-1}$
k_p = thermal conductivity of particles, $W \cdot m^{-1} \cdot K^{-1}$
U_t = particle terminal velocity, $m \cdot s^{-1}$

Greek letters

μ = viscosity of fluidizing gas, $Pa \cdot s$
ρ_g = density of gas, $kg \cdot m^{-3}$
ρ_p = density of particles, $kg \cdot m^{-3}$
τ = contact time of packets, s

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